Optimizing and Controlling the Operation of Heat-Exchanger Networks

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A procedure was developed for on-line optimization and control systems of heat-exchanger networks, which features a two-level control structure, one for a constant configuration control system and the other for a supervisor on-line optimizer. The coordination between levels is achieved by adjusting the formulation of the optimization problem to meet requirements of the adopted control system. The general goal is always to work without losing stream temperature targets while keeping the highest energy integration. The operation constraints used for heat-exchanger and utility units emphasize the computation of heat-exchanger duties rather than intermediate stream temperatures. This simplifies the modeling task and provides clear links with the limits of the manipulated variables. The optimal condition is determined using LP or NLP, depending on the final problem formulation. Degrees of freedom for optimization and equation constraints for considering simple and multiple bypasses are rigorously discussed. An example used shows how the optimization problem can be adjusted to a specific network design, its expected operating space, and the control configuration. Dynamic simulations also show benefits and limitations of this procedure.

Introduction

Operation and control of heat-exchanger networks

The design of heat-exchanger networks (HEN) has been one of the central subjects in the field of thermally integrated chemical processes during the last 15 or 20 years. However, works on control and operation of these systems are not many. The concern on operation and control arises when they have to deal with significant changes in the operating conditions. A flexible network is a system capable of absorbing long-term variations on inlet stream conditions or having the capability of changing stream temperature targets significantly. This concept has been discussed by specialists like Floudas and Grossmann (1986), Calandranis and Stephanopoulos (1986), Kotjabasakis and Linnhoff (1986), Yee and Grossmann (1990), Galli and Cerdá (1991), and Papalexandri and Pistikopoulos (1992), among others.

On the other hand, controllability is associated with short-term perturbations, stability, and safe transitions from one operating point to another. Interaction between design and operability concepts during the synthesis of HENs has been arising as an unavoidable need. The convenience of a careful analysis during the design stage including a preliminary control structure capable of achieving regulation objectives have already been pointed out. The following articles are examples of contributions made in that direction: Marselle et al. (1982), Beautyman and Cornish (1984), Calandranis and Stephanopoulos (1988), Huang and Fan (1992), and Mathisen et al. (1992). Though the amount of work on this subject is becoming important, not much has been said about on-line optimization methods to operate a HEN in an efficient manner and without resigning the temperature control targets assigned to process streams. Lately, important results on this topic were presented in Mathisen's thesis (1994) and related articles.

However, additional contributions are necessary to define all the specific relationships to be included for on-line optimization, a practical procedure to configure the associated regulating control loops and the main clues for coordinating both parts.

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The objectives of this work

This article is aimed at providing a simple modeling representation that gives the basis for on-line steady-state optimization and helps to determine an associated control-loop configuration for any kind of HEN. In this development, it is assumed that all necessary exchangers, utility units, and the connecting structure are completely defined, as well as the heat-transfer areas. The operating space is defined by the steady states used during the design stage, or new ones.

A specific model of the HEN is used for searching the convenient operating point attending to the full network capacity under given inlet stream conditions and temperature targets. Thus, the optimizer determines the best possible set of commands using all the resources available in the network for each working condition at any time. These commanding actions represent load changes for the regulation system that cares for temperature targets.

Simultaneous on-line optimization and regulating control might create conflicting actions. This is particularly true when moving the system from an operating point to another implies setting at full capacity or shutting down one or more units used in the regulation task. In this article, a practical answer is given to the problem by modifying associated statements in the optimization problem.

The organization of this article is as follows: in the following section the process-stream energy balances are written in terms of the energy tasks to be accomplished on the streams and the heat duties to be performed by the individual units. It is also shown that the total energy balance relates the number of independent outlet-free streams with network operability and flexibility for optimization. The operating constraints of heat exchangers and utility units are presented in the third section, where the links with manipulated variables are highlighted. In the fourth section, the optimal operation problem is formulated. The fifth section deals with the configuration of the control system; it gives some clues for selecting manipulated variables, describes the use of control protections in LP problems, discusses the use of slack variables for coordination between both levels, and gives guidelines to configure the regulating control loops. Finally, an application example is analyzed under different conditions in the sixth and seventh sections, where several results of dynamic simulations are also presented. The concluding remarks are given in the eighth section.

Energy Balances of the Network

Stream energy balances

The energy balance for a general hot stream i, in a network can be written as follows:

$$w_i c_i T_i^{\text{out}} = w_i c_i T_i^{\text{in}} - \left(\sum_{k \in K_i} q_k \right) - q_{c_i}; \qquad i \in \mathcal{H}, \quad (1)$$

where $w_i c_i$ is the stream energy capacity per unit time, T_i^{in} is the stream temperature at the network inlet, T_i^{out} is the stream temperature at the network outlet, q_k is the heat transferred at the exchanger k, K_i stands for the set of heat exchangers on the stream path, q_{c_i} is the heat left at a final cooling utility, and $\mathcal{H}$ stands for the whole set of hot streams.

Similarly, for a general cold stream j,

$$w_j c_j T_j^{\text{out}} = w_j c_j T_j^{\text{in}} + \left(\sum_{k \in K_j} q_k \right) + q_{h_j}; \qquad j \in \mathcal{C}, \quad (2)$$

where the adopted nomenclature is similar to that just explained, q_{h_j} being the heat taken at a final heating utility.

The tasks to be performed at any instant on hot process streams are defined as follows:

$$Q_i = w_i c_i T_i^{\text{out}} - w_i c_i T_i^{\text{in}}, \qquad i \in \mathcal{H}, \quad (3)$$

and for cold streams,

$$Q_j = w_j c_j T_j^{\text{out}} - w_j c_j T_j^{\text{in}}, \qquad j \in \mathcal{C}. \quad (4)$$

Notice that these tasks, Q_i and Q_j, take into account all possible process perturbations to the network, that is, changes of inlet stream properties like temperature, flow rate, or heat capacity, and changes in temperature targets. Q_i and Q_j can also be written as functions of the heat duties performed by the heat exchangers and utility units,

$$Q_i = -\sum_{k \in K_i} q_k - q_{c_i}, \qquad i \in \mathcal{H}, \quad (5)$$

and

$$Q_j = \sum_{k \in K_j} q_k + q_{h_j}, \qquad j \in \mathcal{C}, \quad (6)$$

where K_i and K_j stand for all the heat exchangers on streams i and j, respectively.

Equations 5 and 6 show the direct relationship between time-variant tasks to be performed by the network on process streams and the heats exchanged by the individual units that must be accommodated by the control system to satisfy the temperature targets.

Total network energy balance requirements

From an operating point of view, it is convenient to classify cold and hot process streams according to whether they have defined a temperature target and, if so, whether they may use service equipment to reach it. To this end we set the notations

$\mathcal{H}_1$: {hot streams with final temperature targets and without utility units}

$\mathcal{H}_2$: {hot streams with final temperature targets and with utility units}

$\mathcal{H}_3$: {hot streams without final temperature targets}

$\mathcal{C}_1$: {cold streams with final temperature targets and without utility units}

$\mathcal{C}_2$: {cold streams with final temperature targets and with utility units}

$\mathcal{C}_3$: {cold streams without final temperature targets}.

The heat integration achieved in the network implies the equality between the amount of heat given up at the heat exchangers by all the hot streams and the amount of energy received in these units by all the cold streams. In other words, we have

$$\sum_{i \in \mathcal{H}} \sum_{k \in K_i} q_k = \sum_{k \in K} q_k = \sum_{j \in \mathcal{C}} \sum_{k \in K_j} q_k. \tag{7}$$

Then, using Eqs. 3 to 6 we can write the energy balance for the whole network as follows:

$$\sum_{i \in \mathcal{H}_2} q_{c_i} - \sum_{j \in \mathcal{C}_2} q_{h_j} + \sum_{i \in \mathcal{H}_3} w_i c_i T_i^{\text{out}} + \sum_{j \in \mathcal{C}_3} w_j c_j T_j^{\text{out}}$$
$$= \sum_{i \in \mathcal{H}} w_i c_i T_i^{\text{in}} + \sum_{j \in \mathcal{C}} w_j c_j T_j^{\text{in}} - \sum_{i \in \mathcal{H}_1 \cup \mathcal{H}_2} w_i c_i T_i^{\text{out}}$$
$$- \sum_{j \in \mathcal{C}_1 \cup \mathcal{C}_2} w_j c_j T_j^{\text{out}}. \tag{8}$$

In a HEN being monitored at a regular time interval, the variables appearing in the righthand side in Eq. 8 are completely determined, the first two terms represent process stream inlet conditions, and the second two terms include target temperatures for both hot and cold streams, respectively. The variables in the lefthand side (q_{c_i} for $i \in \mathcal{H}_2$, q_{h_j} for $j \in \mathcal{C}_2$, T_i^{out} for $i \in \mathcal{H}_3$ and T_j^{out} for $j \in \mathcal{C}_3$) are variables that can be accommodated to satisfy the conditions imposed by the righthand side terms and/or to optimize (which means reducing the first two terms as much as possible). Thus, those variables in the lefthand side are not completely independent. This dependence can be expressed as follows.

Remark 1 (Total Energy Balance Condition). Let us denote by s the total number of utility units (connected to streams in $\mathcal{H}_2$ and $\mathcal{C}_2$), and by n^o the total number of process streams without final temperature regulation (those in $\mathcal{H}_3$ and $\mathcal{C}_3$). Then, the number

$$f = n^o + s - 1, \tag{9}$$

represents the possibility—under certain operating ranges—of optimization ($f > 0$), just operability ($f = 0$), or impossibility of keeping all the proposed target temperatures ($f = -1$).

Note that f gives the amount of *independent outlet-free streams* in the network, including utilities streams—the -1 comes from the relationship (Eq. 8) itself. If all process streams have temperature targets, it basically measures the possibilities of changing utility duties, while the righthand side terms in Eq. 8 are constant; this means *structural possibilities of optimization* once process stream targets and inputs conditions are fixed. In Appendix A, we analyze different network structures and compare this number with the degrees of freedom defined by Marselle et al. (1982).

Heat-Transfer Operation Constraints

Single heat exchanger

Let us define *energy available at a stream-match* as the maximum heat that can be transferred between two streams. For convenience, we write this as the amount of energy that *the cold stream* takes when heated from T_j^0 to T_i^0, that is,

$$L = w_j c_j \left(T_i^0 - T_j^0 \right), \tag{10}$$

where i stands for hot stream, j stands for cold stream, and the superscript 0 stands for heat-exchanger inlet conditions. Notice that, according to this definition, L is a *nonnegative quantity.* This virtual amount of exchangeable energy is useful for determining the heat transferred at a heat exchanger,

$$q = eL, \tag{11}$$

where $e \in [0, 1)$ is the heat-exchanger efficiency defined by Dodge (1944). It is clear in Eq. 11 that e limits the amount of heat that is transferred when a finite-exchange area is available at the stream match under consideration.

A short inspection of relationships in Appendix B shows that if any of the flow rates, w_i or w_j, varies from a maximum value to zero, the heat-exchanger duty changes in a similar manner, that is,

- If $w_j \to 0$, $L \to 0$, $e \to 1$, and $q \to 0$.
- If $w_i \to 0$, $L \neq 0$ and bounded, $e \to 0$, and $q \to 0$.

These operational limits are found in utility units where the total flow rate of the utility stream is manipulated, or in heat exchangers with a bypass used for temperature regulation or heat-transfer control. In both cases, one stream changes from a maximum value to zero, determining different points in the operation space (Aguilera and Marchetti, 1995).

Remark 2 (Single Unit Constraints). Let o stand for fully open control valve or fully closed bypass. Then the operative interval of a single heat exchanger, where a total flow rate or a bypass ratio is manipulated, can be written in terms of the stream-match available energy L^o, given by Eq. 10 and the heat-exchanger efficiency e^o, as follows:

$$0 \le q \le e^o L^o, \tag{12}$$

where the extremes, $q = 0$ and $q = e^o L^o$ imply fully closed and fully open control valve, or fully open and fully closed bypass, respectively. Any other intermediate condition represents an operating point where the control valve, or the bypass, is partially open.

A corollary of the preceding result is that any intermediate value for the manipulated variable, u, can be estimated through the nonlinear equation

$$q = e(u) L(u) \le e^o L^o. \tag{13}$$

Knowing estimated values of critical manipulated variables during the search of optimal operating points helps to prevent control saturation.

Operative constraints for a heat exchanger in the network

The operation of a heat exchanger that is embedded in a network is constrained by similar relationships, the only different being the way in which the available energy L^o is calculated at each match. This is done by accounting for all the heats previously exchanged by the streams reaching each exchanger in the network. For convenience, the following sets are defined:

$pre_i(k)$ = {Heat exchangers on hot stream i prior to the heat exchanger k},

$pre_j(k)$ = {Heat exchangers on cold stream j prior to the heat exchanger k},

where $i \in \mathcal{H}$, and $j \in \mathcal{C}$. Hence, if for simplicity, no stream split is involved, the temperatures of streams reaching the heat exchanger k can be written as follows:

$$T_i^0(k) = T_i^{\text{in}} - \frac{1}{w_i c_i} \sum_{l_i} q_{l_i}, \qquad l_i \in pre_i(k), \tag{14}$$

$$T_j^0(k) = T_j^{\text{in}} + \frac{1}{w_j c_j} \sum_{l_j} q_{l_j}, \qquad l_j \in pre_j(k), \tag{15}$$

where the superscript 0 stands for the inlet to the exchanger k. Based on Eqs. 14 and 15, the available energy for exchanger k is now written as

$$L_k = w_j c_j\left(T_i^{\text{in}} - T_j^{\text{in}}\right) - R_k \sum_{l_i} q_{l_i} - \sum_{l_j} q_{l_j}, \tag{16}$$

where R_k is the heat-capacity flow-rate ratio $w_j c_j / w_i c_i$ at the corresponding stream match. Hence, the constraints (Eq. 12) can be extended to each heat-transfer unit in a network.

Remark 3 (Network Unit Constraints). The operative interval of a heat exchanger k in a network can be represented by

$$0 \le q_k \le e_k^o L_k^o, \tag{17}$$

where L_k^o is given by Eq. 16 and e_k^o is the heat-exchanger efficiency, both evaluated at full flow-rate values.

Constraint 17 is valid for any process-to-process heat exchanger {1, ne} or utility unit {1, s}. The interpretation of the span and position of the control valve associated with the unit k in a network is similar to what was indicated for an isolated unit. Furthermore, notice that for a heat-exchanger unit without bypass the rightside constraint in Eq. 17 becomes an equality.

Optimal Operation Problem

The optimal operation problem was first defined by Marselle et al. (1982), where both regulation and optimization are discussed. Lately, an interesting procedure by Mathisen et al. (1994) leads to alternative implementations using control logic.

In the approach presented here, the coordination between the optimizer and the regulating system is emphasized. An important difference with Kotjabasakis and Linnhoff (1986) is that all intermediate stream temperatures are avoided by working with heat-exchanger duties almost exclusively. Furthermore, since the network structure and the heat-exchanger areas are completely defined, the formulation is simpler than for the synthesis problem (Colberg and Morari, 1990, for instance). For the same reason, sequential procedures and HEN superstructures are not necessary (see Yee and Grossmann, 1990).

Under this framework, Eqs. 5 and 6 are used as equality constraints since they define the main tasks required of the network. Constraint 17 accounts for the individual capacities of heat-exchangers and utility units. The objective function is formulated in terms of the minimization of utilities or the maximization of the energy integration. For the second case, the optimization problem can be stated as follows (Aguilera and Marchetti, 1995).

Remark 4 (Steady-State Optimization). The optimal operating point of a fully defined heat-exchanger network can be obtained by solving the following maximization problem:

$$\max \sum_k c_k q_k \qquad k \in \{1, ne\}, \tag{18}$$

subject to

$$\begin{aligned} -\sum_{k \in K_i} q_k - q_{c_i} &= Q_i, & i &\in \mathcal{H}, \\ \sum_{k \in K_j} q_k + q_{h_j} &= Q_j, & j &\in \mathcal{C}, \\ q_k &\le e_k^o L_k^o, & k &\in \{1, ne+s\}, \\ -q_k &\le 0 & k &\in \{1, ne+s\}. \end{aligned} \tag{19}$$

Note that if there are no stream splits in the network, the preceding will be an LP optimization problem. This is because $e_k^o L_k^o$, $k \in \{1, ne+s\}$, are linear relationships among the variables q_k, with constant coefficients calculated from real-time data at selected time periods. However, if the optimizer has to search for optimal split ratios, some constraints become nonlinear functions of changing flow rates due to changing splits, that is, in general we may write

$$q_k \le e_k^o(x_s) L_k^o(x_s), \tag{20}$$

where x_s is the split fraction.

Remark 5 (Multiple Bypasses). Bypasses made over more than one unit produce nonlinear bounds similar to Eq. 20 for every constrained unit following the first one after the stream split. Furthermore, since an estimation of the bypass fraction x_k is needed, the constraint for the first unit after a multiple-bypass split must become an equality,

$$q_k = e_k(x_k) L_k(x_k) \le e_k^o L_k^o. \tag{21}$$

In any case, numerical methods are available to provide a solution of the resulting NLP problem, which can be used for an on-line search of the optimal operating point. Regarding convergence, recall that split and bypass ratios belong to the [0, 1] interval, that the efficiency is a smooth function with values between 0 and $e^o(<1)$, and that a good initialization for these NLP problems comes from associated LP solutions obtained for the same network with single bypasses only and selected values of stream-split ratios. Though these arguments do not prove that a global optimum is achieved, we can expect that for on-line work the solution obtained at time t is always a *good initial guess* to increase the possibilities of holding the best operating condition at $t + \Delta t$.

The weights used in the objective function (Eq. 18) might have just a formal meaning; however, if minimum utility is required, that is,

$$\min \sum_r c_r q_r \qquad r \in \{1, s\}, \tag{22}$$

a selective importance can be given to the different utilities according to its availability or actual cost, or any other particular condition. For instance, since the LP (or NLP) problem might have several (or infinite) solutions with the same utility consumption, additional features could be conveniently included in the formulation.

Remark 6 (Objective Functions for Constant Control Configuration). The objective function should denote preferences and decisions made during the control configuration in order to move the system to the closest operating point and reduce load disturbances caused by the optimizer itself.

This means that, in some cases, cost coefficients in the objective function should be arbitrarily adopted to favor the closest solution from a set of equally valued alternatives. The example discussed in the sixth and seventh sections should make this point clearer.

Control-System Configuration

Several articles dealing with control configurations in HENs have been published in the last few years. Calandranis and Stephanopoulos (1988) suggested a structural approach based on the identification and analysis of the routes that energy disturbances follow through the network. Mathisen et al. (1992), used controllability measures to select bypasses and appropriate pairings. More recently, Mathisen et al. (1994) proposed that selection of the heat exchangers to be bypassed must be done by following the signs of the elements of a transfer-function matrix.

In this article we analyze the use of a two-level control structure, as shown in Figure 1. The first level is for a supervisor system that works by optimizing the operation and/or protecting, whenever possible, process-stream control loops from steady saturations in manipulated variables. The second level or plant level consists of two parts: (1) a conventional multiloop regulating system to keep process streams on their targets, and (2) a complementary control system for tracking supervisor commands, that is, to drive the network to the desired steady state. Notice that the supervisor commands work like load changes or disturbances for the regulating system. Both levels must be carefully defined and coordinated to be able to hold optimal operating conditions without losing the final temperature targets.

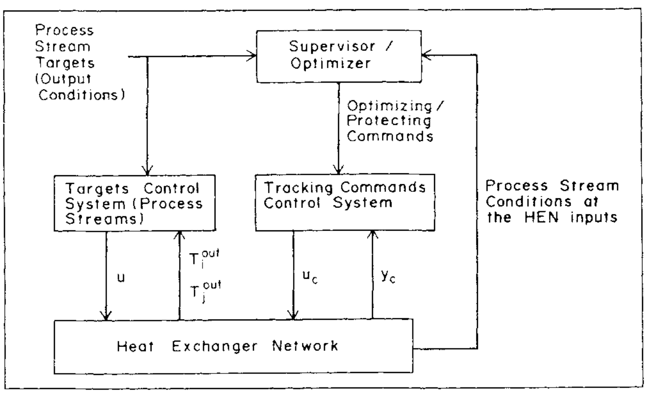

Figure 1. Two-level control structure.

Regarding the second level, this article focuses mainly on configuring the process-stream control system so as to obtain the best possible control performance. However, a few words need to be said with regard to configuring the tracking-commands control system. First, in this approach the determination of which variables will be manipulated by the tracking system depends on how the regulating system is configured. Second, observe that the solution of the optimization problem defined in the previous section is expressed mainly in terms of heat loads. Therefore, the commands coming from the supervisor basically define desired heat loads (q_{set}) or split ratios not defined by the process-stream control system. Controlling a heat-exchanger heat load with a bypass is quite straightforward since it reduces to controlling the temperature of the bypass stream—the flow rate must be available for the supervisor at each sampling instant anyway. Hence, setpoint commands are computed simply by

$$T_{set} = T^0 \pm \frac{q_{set}}{wc}.$$

From a steady-state economic point of view it is irrelevant whether the hot (−) or the cold (+) stream is bypassed. Moreover, the design of these controllers should follow similar rules as those for bypass temperature control of final process streams (Rotea and Marchetti, 1997). Alternatively, the supervisor manipulates utility streams when commanding service units; in this regard, several years ago Shinskey (1977) discussed heat-load control on heat exchangers using appropriate instrumentation.

Manipulated variables for temperature control

Rotea and Marchetti (1997) presented a detailed analysis of temperature control in heat-exchanger-plus-bypass systems. They have shown that the configuration for the best possible temperature control has a bypass on the stream whose temperature is to be controlled. When such configuration is used, a simple integral controller suffices for solving performance and stability problems. These claims have been supported with analytical results and nonlinear simulations. All the steps of their analysis and conclusions can be extended to problems with direct-effect bypasses over two or more units.

Heat exchangers, utility units, and stream splits are the three basic sources of manipulated variables in a HEN. Following Mathisen et al. (1992) and Rotea and Marchetti (1997), direct-effect bypasses are preferred manipulated variables. Mathisen (1994) supports the same conclusion, but also points out that it might be advisable to bypass the uncontrolled stream when this has lower heat-capacity flow rate.

Thus, as a starting point it is assumed that each heat exchanger in the network provides a bypass that can be used for regulation. Even though permanent utility units can be partially bypassed by the process streams, this alternative is not used, because it implies a good but costly regulation since the utility stream has to be larger than that required by the heat duty. Hence, manipulated variables associated with utility units are utility stream flow rates, and the benefit in performance comes from the fact that they are typically final units on controlled process streams. Alternatively, splits increase interaction and, in the general case, they do not show any particular convenience for regulating control.

Coordination between regulating and supervisor levels

The classic loop-configuration problem in process control comes from the need for pairing controlled variables with manipulated variables, so as to obtain the best performance. Typically in this problem, manipulated variables are continuously available for the regulation task. However, in a flexible HEN that must work optimally in a given operating space, the optimizer commands might lead to the saturation of manipulated variables, leaving the system partially uncontrolled. This is a consequence of heat exchangers or utility units that should go out of operation or should work at full capacity to satisfy the optimal conditions.

For instance, the minimum utility requirement might ask a utility to cease operating, thus creating a conflicting situation where the regulating system needs the control valve open to keep the closed-loop control working, while the optimizer tries to close it down. The problem is solved using proper limit values for the duties of units supporting manipulated variables that might saturate (Aguilera and Marchetti, 1995). This also means that some energy integration has to be sacrificed in order to preserve regulation.

In cases where a unit used for regulation is indirectly driven out of operation by the optimizer, a protection must be created by modifying the nonnegativity conditions in Eq. 19, as follows:

$$-q_k \leq -\alpha_k, \qquad k \in \{1, ne+s\}, \tag{23}$$

where α_k is the resigned amount of energy determining how close to saturation the manipulated variable can reach.

The constraints might also have to be modified for underdesigned units. For instance, a bypass used for regulation might close completely trying to keep the controlled variable at the setpoint. In cases like this, the constraints to be modified are those corresponding to maximum heat-transfer capacity,

$$q_k \leq \beta_k e_k^o L_k^o, \qquad k \in \{1, ne+s\}, \tag{24}$$

where β_k is a fraction close to 1, reducing the amount of energy that can be transferred in unit k.

The importance of writing the protections as previously indicated, is that otherwise using explicit limits for the manipulated variables generates nonlinear relationships in the optimization problem.

Remark 7 (Control Protections in LP Problems). If the HEN stream pattern allows an LP optimization problem, protections for a manipulated variable, u_k, can be formulated using Eqs. 23 and 24 without losing the linear condition; once the values for minimal and maximum operation, $\underline{u}_k$ and $\overline{u}_k$, are adopted, the parameters α_k and β_k can be estimated by

$$\alpha_k = e_k(\underline{u}_k) L_k(\underline{u}_k), \tag{25}$$

$$\beta_k = \frac{e_k(\overline{u}_k) L_k(\overline{u}_k)}{e_k^o L_k^o}. \tag{26}$$

Also, final adjustments can be decided by operators based on process needs and experience. The meaning of the parameters α_k and β_k is easy to comprehend since they weigh the importance of keeping a final temperature target against heat integration.

The use of protections to prevent steady-state saturations of manipulated variables is a major part of the coordination between optimizing and regulating control systems. In this regard, slack variables typically used in the numerical solutions of optimization problems are very helpful. They basically measure the distance of each inequality constraint to become active. Inspection of these variables in the solution to the optimization problem is very important in determining which unit constraint in Eqs. 19 must be modified as indicated earlier. Observe that a slack variable greater than zero means the bypass is partially open in a heat exchanger or a control valve is partially closed in a utility.

Remark 8 (Using Slack Variables to Determine Control Protections). Units under limit conditions in a HEN operation are detected through slack variables equal to zero in the optimal solution. A slack variable remaining equal to zero in the whole operating space and associated with a heat exchanger not used for regulation indicates that the bypass is not necessary.

Regulating control loops and optimizer commands

As with any other multivariable system, the definition of a multiloop control structure takes one manipulated variable per controlled variable. Control performance in HENs mainly depends on the kind of manipulated variables associated with the feedback loops. Hence, if the goal is the best possible control quality for the largest amount of process streams, we suggest defining the control configuration by selecting manipulated variables according to the performance they can give: (1) direct bypass on controlled streams; (2) service flow rates —typically final equipment on process streams; (3) nondirect action bypasses on internal exchangers; and (4) stream-split ratios. For this sequence, achievable control-loop performance ranks from near perfect control for the first one (Rotea and Marchetti, 1997) to sluggish and interacting for the last one.

From the control point of view, each bypass is associated with the heat duty of the first heat exchanger after the two-way valve, even though it jumps over more than one unit. A direct bypass jumping above two or three units does not reduce its capacity for control performance when compared to jumping over a single unit. However, multiple bypasses might show limitations, particularly on the operating range of heat exchangers having both outlet temperatures controlled (Mathisen, 1994). Also, large physical distances might adversely affect the piping cost, but this is very much a layout-dependent problem to be balanced with other costs during the design stage.

Configuring control loops for flexible HENs could be a complicated task, depending on the network structure. Many times it implies going back and forth with decisions until all the eventual operating conditions are properly covered. However, our first goal must be to reach *maximum control performance* for the given network without losing heat integration. Other considerations like piping cost, larger or smaller operability space, practicality of the solution, etc., come in a second step when we know how much control quality is sacrificed and on which control target. Once the regulating system is defined, the remaining manipulated variables become optimizing or protecting commands.

The following steps help us to move forward in a rational and orderly manner toward the complete definition of the feedback loops for the regulating system and the coordination with the supervisor level:

1. Compute the number of independent outlet-free streams, f, using Eq. 9 to have a preview of structural flexibility and to confirm that using on-line optimization may be profitable for the network at hand.
2. Set the convenient objective function.
3. Get the specific optimization model following the pattern given by constraints 19. The first time they can be written assuming a single bypass at each process–process exchanger so as to reach an LP formulation. Constraint equations for service equipment known *a priori* to be overdesigned need not be written.
4. Solve the optimization problem for nominal operating conditions—the steady state in which the system will be found most of the time. Use the optimization problem like a steady-state simulator to examine the main points in the expected operating space in order to forecast different operating conditions (do this the first time for $\alpha_k = 0$ and $\beta_k = 1 \forall k$). From the results determine temporary units and units that might work at full capacity under some network condition.
5. Pair controlled variables with bypasses to closest heat exchangers on the controlled streams. Prefer direct bypasses associated with permanent overdesigned heat exchangers. Whenever possible, avoid final configurations that have heat exchangers under two bypasses.
6. Complete the loop configuration pairing the remaining controlled variables to utility streams. A decision must be made at this point for process streams with a temporary or underdesigned heat exchanger (not assigned for regulation yet), followed by a permanent utility; the bypass to the heat exchanger is the proper choice if control performance is important, but the utility flow rate must be adopted if energy integration is prioritized.
7. Determine all the manipulated variables to be commanded by the optimizer. Typically these are stream splits, temporary utilities, and process–process exchangers far from final temperature targets.
8. If necessary, modify unit constraints in the original optimization problem to consider any change in bypass configuration or to protect the regulating system from steady saturation of manipulated variables.
9. Repeat the simulations with proper α_k and β_k values, and review the whole configuration, making sure of its operability in the desired operating space.

This procedure leads to practical and effective loop configurations that are coordinated with the supervisor level. However, there is no claim here that the resulting control configuration is optimal in some sense. The trade-off between control performance and utility consumption is a difficult problem, and no rule seems to be completely general. Even for a HEN with its structure and areas defined, control configuration depends a great deal on target priorities and the desired operating window.

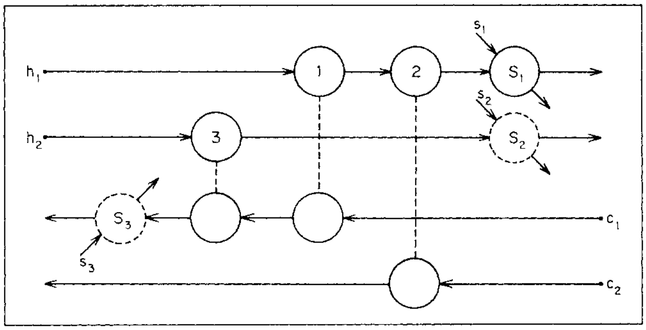

Figure 2. Typical synthesis diagram for the example problem.

Application Example

Figure 2 shows an illustrative example problem where supervisory control is convenient. Though there are many networks that could be taken for a demonstration, this is a simple, interesting case that suffices to define alternative optimization problems and their related control configurations. The existence of such alternatives for a simple network like this is an indication that the definition of the optimization problem and the corresponding control configuration depend on each particular application.

Based on the nominal stream conditions given in Table 1, it is found that a minimum of three heat exchangers (see Figure 2) and one utility unit (S_1 or S_2) are needed for maximum heat integration in this problem. However, two more utilities are included to extend the operating space, providing flexibility to absorb changes that otherwise cannot be afforded (Marselle et al., 1982); it should be intuitively clear that the *base network*, consisting of four units, is operable in a reduced space around the nominal point, mainly due to slight overdesign of the heat-transfer areas and operative displacements in favorable directions.

In this example, there are no streams without final temperature regulation, $n^o = 0$; there are three utility units, $s = 3$, and consequently Eq. 9 indicates that the general energy balance leaves $f = 2$ independent outlet-free streams, allowing the possibility of using an optimum criterion to fix the operating point.

The units needed under nominal conditions provide the first clue for control configuration. Hence, additional arguments should be found to decide which utility, S_1 or S_2, will

Table 1. Stream Conditions for the Nominal Operation Point

Stream	T^{in} (°C)	T^{out} (°C)	$w^o c$ (kW/°C)
h_1	90	40	50
h_2	130	100	20
c_1	30	80	40
c_2	20	40	40
s_1	15	—	35 (max.)
s_2	30	—	30 (max.)
s_3	200	—	10 (max.)

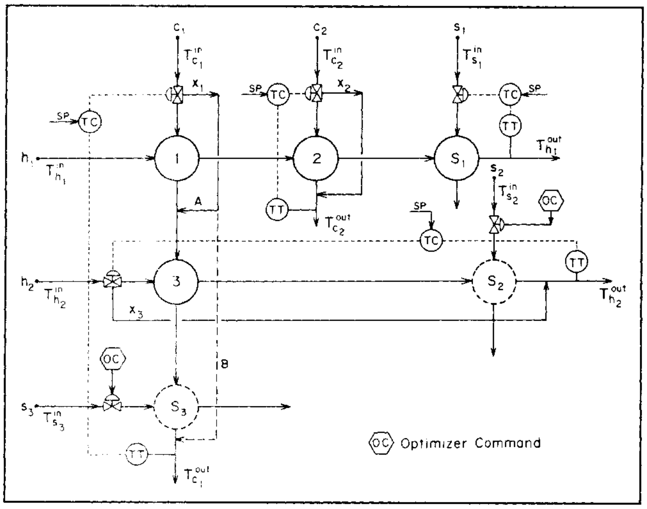

Figure 3. Control loop configuration for the network of Figure 2.

be working under this nominal point. Let us assume that S_1 deserves such a privilege, perhaps because utility stream s_2 is more expensive than s_1, or because s_2 is a source of frequent disturbances, while the first one gives more chances of steady conditions. If this is the case, final units S_2 and S_3 on streams h_2 and c_1, respectively, are temporary units that cannot be selected for temperature regulation. Notice that keeping them working at the network nominal point means increasing the minimum utility level just for controlling reasons. Hence, the network shows here an interesting and quite frequent problem: heat exchanger 3 becomes a single unit, with two outlet streams having regulation targets when neither S_2 nor S_3 is working. On the other hand, the alternative selection of S_2 as the permanent unit instead of S_1 leads to a configuration where exchanger 2 is a single unit with two controlled outputs.

Figure 3 shows two control configurations, one that uses a direct-bypass (dashed line), prioritizing control performance on the final temperature of stream c_1, and the other that avoids doubly bypassing heat exchanger 3. In both cases, units S_2 and S_3 are commanded by the optimizer according to the needs of the operating point.

Regulating $T_{c_1}^{out}$ with a single bypass around exchanger 1

We know that regulating $T_{c_1}^{\text{out}}$ with a single bypass around exchanger 1 will have lower control performance than when doing it by a direct multiple bypass. However, if $T_{c_1}^{\text{out}}$ has to be maintained within specified bounds and the setpoint value is the expected average temperature, this configuration would be acceptable. Taking the network to an optimal operating point implies knowing the better combination of heat duties, q_k, at exchangers 1 to 3, that produces the largest heat integration. Modeling equations proposed in previous sections lead to the formulation of the following LP problem:

$$\max\{q_1 + q_2 + q_3\}, \tag{27}$$

satisfying the stream requirements,

$$\begin{aligned} -q_1 - q_2 - q_{S_1} &= Q_{h_1}, \\ -q_3 - q_{S_2} &= Q_{h_2}, \\ q_1 + q_3 + q_{S_3} &= Q_{c_1}, \\ q_2 &= Q_{c_2}, \end{aligned} \tag{28}$$

the equipment or unit constraints,

$$\begin{aligned} q_1 &\le e_1^o L_1^o, \\ q_2 &\le e_2^o L_2^o, \\ q_3 &\le \beta_3 e_3^o L_3^o, \\ q_{S_1} &\le \beta_{S_1} e_{S_1}^o L_{S_1}^o, \end{aligned} \tag{29}$$

and the nonnegative conditions,

$$\begin{aligned} -q_1 \le 0, \quad & -q_2 \le 0, \quad -q_3 \le 0, \\ -q_{S_1} \le -\alpha_{S_1}, \quad & -q_{S_2} \le 0, \quad -q_{S_3} \le 0. \end{aligned} \tag{30}$$

The right side of Eqs. 28 are defined by Eqs. 3 and 4, and the energy available at the stream matches are given by

$$\begin{aligned} L_1^o &= w_{c_1}^o c_{c_1}\left(T_{h_1}^{\text{in}} - T_{c_1}^{\text{in}}\right), \\ L_2^o &= w_{c_2}^o c_{c_2}\left(T_{h_1}^{\text{in}} - T_{c_2}^{\text{in}}\right) - R_2^o q_1, \\ L_3^o &= w_{c_1}^o c_{c_1}\left(T_{h_2}^{\text{in}} - T_{c_1}^{\text{in}}\right) - q_1, \\ L_{S_1}^o &= w_{s_1}^o c_{s_1}\left(T_{h_1}^{\text{in}} - T_{s_1}^{\text{in}}\right) - R_{S_1}^o (q_1 + q_2). \end{aligned} \tag{31}$$

In this formulation, utility units S_2 and S_3 are assumed to have enough heat-transfer area, and the associated utility temperatures to be far from process-stream temperatures; this implies that constraints for these two units are not necessary since they will never be active. Table 2 gives the values of R_k^o

Table 2. Heat-Exchanger Parameters

	AU	LP Problem			Design A		Design B	
k	(kW/°C)	R_k^o	e_k^o	F_T	R_k	e_k	R_k	e_k
1	80	0.8	0.711	0.906	$0.8(1-x_1)$	$e_1(x_1)$	$0.8(1-x_1)$	$e_1(x_1)$
2	50	0.8	0.588	0.936	0.8	0.568	0.8	0.568
3	20	2.0	0.282	0.930	$2/(1-x_3)$	$e_3(x_3)$	$R_3(x_1, x_3)$	$e_3(x_1, x_3)$
S_1	30	0.7	0.495	0.843	0.7	0.447	0.7	0.447
S_2	20	1.5	0.362	0.930	$1.5/(1-x_3)$	$e_{S_2}^o(x_3)$	$1.5/(1-x_3)$	$e_{S_2}^o(x_3)$
S_3	10	4.0	0.149	1.000	4.0	0.149	4.0	0.149

and e_k^o for nominal stream conditions at completely closed bypasses and maximum utility flow rates. After a few runs, inspection of slack variables at different points belonging to the desired operating space shows that S_1, used for regulation of $T_{h_1}^{\text{out}}$, must be protected to avoid the complete use of its cooling capacity ($\beta_{S_1} \neq 1$) and to avoid turning it off ($\alpha_{S_1} \neq 0$). Similarly, a protection must be included in the constraint of heat exchanger 3 to keep the bypass controlling $T_{h_2}^{\text{out}}$ from closing completely: $\beta_3 \neq 1$.

Recall important features allowing this to be an LP formulation: (1) there are no split streams; (2) there are no constrained units following the first one under a bypass; when S_2 is working, the bypass controlling $T_{h_2}^{\text{out}}$ becomes a multiple bypass, but the utility unit is not constrained; (3) protections from control saturations are made on associated heat duties only, that is, they are not explicit manipulated variable constraints.

Regulating $T_{c_1}^{out}$ with a direct multiple bypass

Let us see how this problem formulation changes in the case where a direct bypass for regulating $T_{c_1}^{\text{out}}$ is decided. The energy performance index (Eq. 27) and the definition of network duties (Eq. 28) remain unchanged; however, heat exchanger 3 is now a constrained unit, following the first one under a multiple bypass. Since the operative capacity of heat exchanger 3 depends on the changing bypass fraction x_1, the third constraint in Eq. 29 becomes nonlinear

$$q_3 \leq \beta_3 e_3^o(x_1) L_3^o(x_1). \tag{32}$$

The functional dependency of e_3^o and L_3^o from x_1 can be made explicit using the relationships shown in Appendix B. For this case, the cold-stream flow rate of heat exchanger 3 is $(1-x_1)w_{c_1}^o$. Notice, however, that the superscript o is kept at the left side of Eq. 32 to indicate that this maximum value is computed for $x_3 = 0$, that is, for bypass of heat exchanger 3 completely closed. Furthermore, x_1 must be known to be used in Eq. 32; as explained at the end of the fourth section, this is done by transforming the inequality constraint of heat exchanger 1 into the equation

$$q_1 = e_1(x_1) L_1(x_1). \tag{33}$$

Again, the relationships in the Appendix B give the basis for writing the detailed expression of Eq. 33.

Hence, when a direct bypass is used for regulating $T_{c_1}^{\text{out}}$, the new system of stream-matched constraints is

$$\begin{aligned} q_1 &= e_1(x_1) L_1(x_1), \\ q_2 &\leq e_2^o L_2^o, \\ q_3 &\leq \beta_3 e_3^o(x_1) L_3^o(x_1), \\ q_{S_1} &\leq \beta_{S_1} e_{S_1}^o L_{S_1}^o. \end{aligned} \tag{34}$$

Besides, since the nonlinear nature of this problem requires limits for x_1, they can be used to provide explicit limits for control protection, or

$$\begin{aligned} x_1 &\leq x_1^{\max}, \\ x_1 &\geq x_1^{\min}. \end{aligned} \tag{35}$$

Multiple bypass and limited size of utility S_2

Let us assume now that the size of the cooler S_2 makes it necessary to include the corresponding unit constraint so that the optimizer be able to determine how large the heat duty that can be safely assigned to it is. Since this utility unit is located under the bypass on stream h_2, there are now two multiple bypasses coupled by heat exchanger 3. The upper bound of q_{S_2} is dependent on the bypass ratio x_3, but must be computed for maximum utility flow rate w_{s_2}. This requires x_3 to be a known value, and consequently the inequality for equipment 3 is transformed into an equality constraint. In other words, the system of unit constraints is now written as follows:

$$\begin{aligned} q_1 &= e_1(x_1) L_1(x_1), \\ q_2 &\leq e_2^o L_2^o, \\ q_3 &= e_3(x_1, x_3) L_3(x_1, x_3), \\ q_{S_1} &\leq \beta_{S_1} e_{S_1}^o L_{S_1}^o, \\ q_{S_2} &\leq e_{S_2}^o(x_3) L_{S_2}^o(x_3), \end{aligned} \tag{36}$$

where the available energies are given by

$$\begin{aligned} L_1(x_1) &= (1-x_1) w_{c_1}^o c_{c_1}\left(T_{h_1}^{\text{in}} - T_{c_1}^{\text{in}}\right), \\ L_2^o &= w_{c_2}^o c_{c_2}\left(T_{h_1}^{\text{in}} - T_{c_2}^{\text{in}}\right) - R_2^o q_1, \\ L_3(x_1, x_3) &= (1-x_1) w_{c_1}^o c_{c_1}\left(T_{h_2}^{\text{in}} - T_{c_1}^{\text{in}}\right) - q_1, \\ L_{S_1}^o &= w_{s_1}^o c_{s_1}\left(T_{h_1}^{\text{in}} - T_{s_1}^{\text{in}}\right) - R_{S_1}^o(q_1 + q_2), \\ L_{S_2}^o(x_3) &= w_{s_2}^o c_{s_2}\left(T_{h_2}^{\text{in}} - T_{s_2}^{\text{in}}\right) - R_{S_2}(x_3) q_3. \end{aligned} \tag{37}$$

Notice that for this particular case, L_3 is a function of x_1 alone. This is a consequence of how L_k is defined in Eq. 16 and the particular position of unit 3 in the network. We have preferred to keep the nomenclature $L_3(x_1, x_3)$ to emphasize the functionality in the general case. Besides, all the efficiencies in Eq. 36 are written as defined in Appendix B; precautions must be taken regarding flow rates only. Efficiency $e_3(x_1, x_3)$, for instance, uses $w_{c_1} = (1-x_1)w_{c_1}^o$ for the cold stream, and $w_{h_2} = (1-x_3)w_{h_2}^o$ for the hot stream. Furthermore, since a prediction of x_3 is now available, the factor β_3 has been eliminated from this constraint because a direct protection on the control variable can now be included,

$$\begin{aligned} x_3 &\leq x_3^{\max}, \\ x_3 &\geq x_3^{\min}. \end{aligned}$$

Simulation Results

An interactive dynamic simulator of HENs developed in INTEC has been used to test the analyzed example. Preliminary characteristics of this simulator were presented by Correa (1994); the simulator is based on a nonlinear model of shell-and-tubes heat exchangers previously reported by Correa and Marchetti (1987).

Table 3. Optimum Steady States Using *Design A*

Case	q_1 (kW)	q_2 (kW)	q_3 (kW)	q_{s_1} (kW)	q_{s_2} (kW)	q_{s_3} (kW)	x_1	x_3	J_{1+2+3} (kW)
Nominal	1,400	800	600	300	0	0	0.290	0.340	2,800
Sequence 1:									
$T_{h_1}^{in} = 80°C$	1,185	800	600	15	0	215	0.272	0.441	2,585
$T_{h_2}^{in} = 140°C$	1,185	800	800	15	0	15	0.272	0.216	2,785
$T_{c_1}^{in} = 40°C$	800	800	800	400	0	0	0.446	0.207	2,400
Sequence 2:									
$T_{c_1}^{out} = 70°C$	1,172	800	428	528	172	0	0.464	0.671	2,400
$T_{c_2}^{out} = 45°C$	1,000	1,000	600	500	0	0	0.564	0.502	2,600
$T_{h_2}^{out} = 90°C$	972	1,000	628	528	172	0	0.579	0.469	2,600

Two designs, represented schematically in Figure 3, are dynamically tested: *design B* is the last case analyzed in the previous section, that is, multiple bypass on stream c_1 with a limited transfer area in utility unit S_2 and, *design A* is the same case but with a single bypass to exchanger 1 on stream c_1, that is, the third relationship in Eq. 36 is a function of x_3 only (see Table 2). This allows a fair comparison, since the same directly manipulated protections can be assigned to both cases. Though the difference between these designs is in the type of bypass to exchanger 1 only, it highlights the problem of operability vs. control performance or, utility consumption vs. control performance.

In this example problem, control configurations of *designs A* and *B* are based on the preference of utility stream s_1 against s_2. Hence, the corresponding NLP problems minimize the objective function:

$$J_s = q_{S_1} + c_{S_2} q_{S_2} + q_{S_3},$$

where $c_{S_2} > 1$, to avoid moving the system around equivalent energy integration points.

For the sake of simplicity, all stream heat capacities per unit mass are taken equal to 4.18 kJ/kg·°C. For this demonstration example we adopt 0.1 and 0.9 as bounds for protecting bypasses from saturation and, $\alpha_{S_1} = 15$ kW (5% of the nominal duty) and $\beta_{S_1} = 0.95$ for protecting the control loop associated with the cooler S_1.

Tables 3 and 4 show NLP results for *designs A* and *B*, respectively, for different steady operating conditions. The first column identifies the nominal steady state and two sequences of changes: (1) the sequence of step disturbances on the inlet stream temperatures; (2) the sequence of step changes on the temperature control targets. Steady-state conditions at a given time instant have proved to be an effective initialization to obtain the NLP solution for the next updating time.

The first row in these tables indicates how heat exchanger duties are selected by the supervisor to reach optimal operation for nominal conditions in Table 1. It also gives the predicted bypass ratios x_1 and x_3. Though for the nominal point both designs yield the same energy integration, the distribution of heat duties is different due to the bypass to exchanger 1. Since when using *design A* heat exchanger 3 receives the whole stream, c_1, the bypass ratio x_3 is larger than that of the steady state of *design B*, where x_3 hits the lower bound 0.1 giving a maximum of 533 kW for q_3. This explains why the optimizer activates the cooler S_2 in *design B*. This result is in agreement with that of Mathisen (1994) in that a multiple bypass adversely affects the operability of a single unit with two output controlled streams.

Let us now analyze the sequences of changes. Recall that we assume that this optimizer receives information about temperature and flow rate of all process streams entering the network, and about all the setpoints defining temperature targets. In other words, the optimizer works attending to changes of the HEN *boundary streams*. The frequency at which the optimizer is activated depends very much on time constants and disturbances associated with the problem. Once the new optimal operating point is determined, the supervisor moves the network through a set of commands *representing load changes for the regulating system*. In our example, the optimizer checks the information and updates the optimal condition by commanding units S_2 and S_3 with a one-minute time interval.

Table 4. Optimum Steady States Using *Design B*

Case	q_1 (kW)	q_2 (kW)	q_3 (kW)	q_{s_1} (kW)	q_{s_2} (kW)	q_{s_3} (kW)	x_1	x_3	J_{1+2+3} (kW)
Nominal	1,467	800	533	233	67.2	0	0.228	0.100	2,800
Sequence 1:									
$T_{h_1}^{in} = 80°C$	1,185	800	596.5	15	3.5	218.5	0.272	0.100	2,581.5
$T_{h_2}^{in} = 140°C$	1,185	800	697	15	103	118	0.272	0.100	2,682
$T_{c_1}^{in} = 40°C$	930	800	670	270	130	0	0.297	0.100	2,400
Sequence 2:									
$T_{c_1}^{out} = 70°C$	1,172	800	428	528	172	0	0.464	0.124	2,400
$T_{c_2}^{out} = 45°C$	1,169.5	1,000	430.5	330.5	169.5	0	0.466	0.100	2,600
$T_{h_2}^{out} = 90°C$	1,169.5	1,000	430.5	330.5	369.5	0	0.466	0.100	2,600

Remark 9. All the controllers actuating through direct bypass on the controlled stream are pure integral controllers, adjusted using the following tuning formula (Rotea and Marchetti, 1997):

$$T_I = 1.353 K_1 T_d, \tag{38}$$

where T_I is the integral constant, T_d is the transport delay between the bypass mixing point and the sensor location, and K_1 is the initial or lead process gain.

The gain value to be used in Eq. 38 is easily calculated by the difference in temperatures of the streams reaching the correspondent match,

$$K_1 = T_{bp}^0 - T_{nbp}^0, \tag{39}$$

where *bp* and *nbp* stand for bypassed and nonbypassed streams, respectively. For multiple bypass control (there are more than one *nbp* stream) the largest difference is conservatively taken. The controller of $T_{h_1}^{out}$, which manipulates the utility stream s_1, and the controller of $T_{c_1}^{out}$, actuating through the single bypass to exchanger 1 in *design A*, are adjusted using Ziegler–Nichols settings based on the process-reaction-curve method.

The simulation experiences for testing the system for temperature disturbances assume that the optimizer detects the input change one minute after it happens (the worst case) and immediately produces a new set of commands. Figures 4 and 5 show dynamic responses obtained from *designs A* and *B*, respectively, when the following sequence of changes is made: after running at the nominal operating point, $T_{h_1}^{in}$ changes at time $t = 0$ from 90°C to 80°C; ten minutes later $T_{h_2}^{in}$ goes from 130°C to 140°C and, after other ten minutes $T_{c_1}^{in}$ changes from 30°C to 40°C.

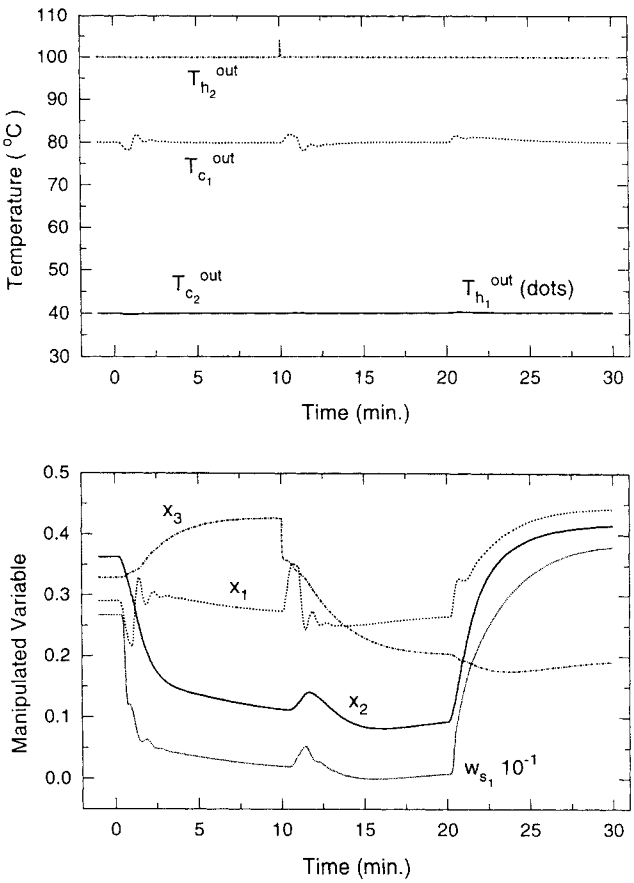

Figure 4. Dynamic response of *design A* to the sequence of temperature step changes on inlet streams.

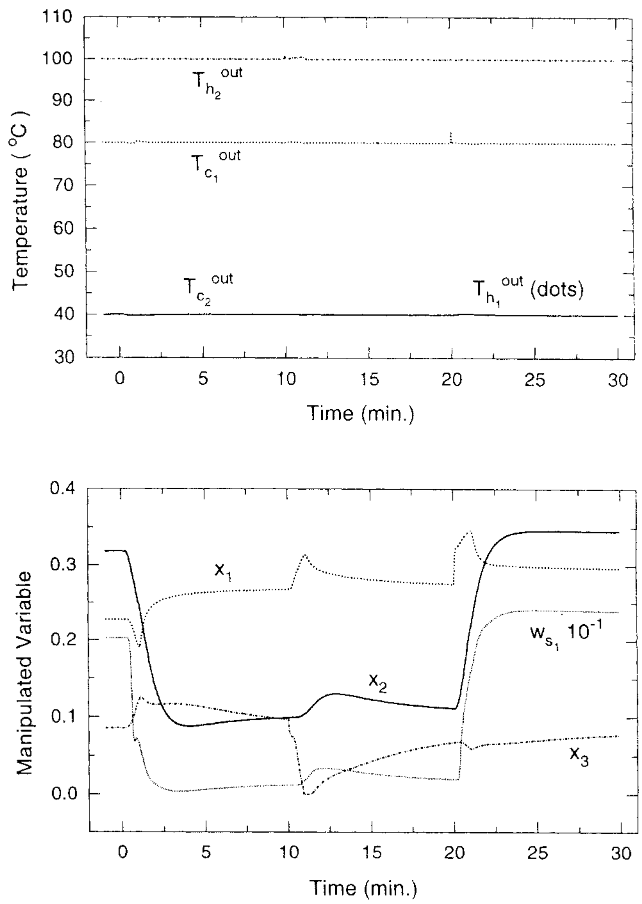

Figure 5. Dynamic responses of *design B* to the sequence of temperature step changes on inlet streams.

After the first load change, new steady states determined for both designs retain the minimum value of q_{S_1} (15 kW) adopted to keep S_1 working for regulation. *Design A* does not use any other manipulated protection for this operating point. However, the result for *design B* shows that protection for x_3 is also active (= 0.1). From the dynamic side, the lags along stream h_1 attenuate the change leaving the controller of $T_{h_1}^{out}$ with enough reaction for almost a complete load rejection. Compensating control actions are mainly observed on manipulated w_{s_1} and x_2. Part of the perturbation goes also to $T_{c_1}^{out}$, which moves x_1 and x_3 to keep related temperatures on target. The effect of the optimizer commands is observed after one minute through the change of tendency in x_1, mainly due to the activation of S_3 from 0 to 215 kW for *design A* and to 218.5 kW for *design B*.

The following load temperature changes can be analyzed in a similar fashion. Note that most of the *internal network dynamics* is observed by inspection of the movements of the manipulated variables rather than the controlled ones. The disturbance rejection is remarkable, as expected, using direct bypass on controlled streams; only a rapid change immediately compensated for is observed, since the temperature disturbance travels through the bypass directly to the mixing point. The height of the pick of the controlled variable depends on the bypass fraction, and it lasts for a time interval similar to the associate closed-loop time delay (Rotea and Marchetti, 1997).

Notice in Figures 4 and 5 that to cope with the first two load temperature changes, w_{s_1} goes *steadily* near saturation; this indicates that α_{S_1} might have to be selected higher than 15 kW, or eventually, that a direct bound for this manipulation should be included in the optimization problem. Some statements must be made regarding the accuracy of predicting stationary values of manipulated variables and therefore to prevent *permanent* saturations: the HEN model used by the optimizer (based on relationships given in Appendix B) will most of the time show differences with the actual plant. Hence, a sort of robust optimization problem arises: as the mismatch increases, so does the error of the predictions of the conditions under which saturations arise.

Remark 10 (Optimization Model Mismatch). The accuracy of the optimization model in predicting conditions of permanent control saturation can be greatly improved by adjusting the individual factors F_T. This is done by solving the nonlinear equation

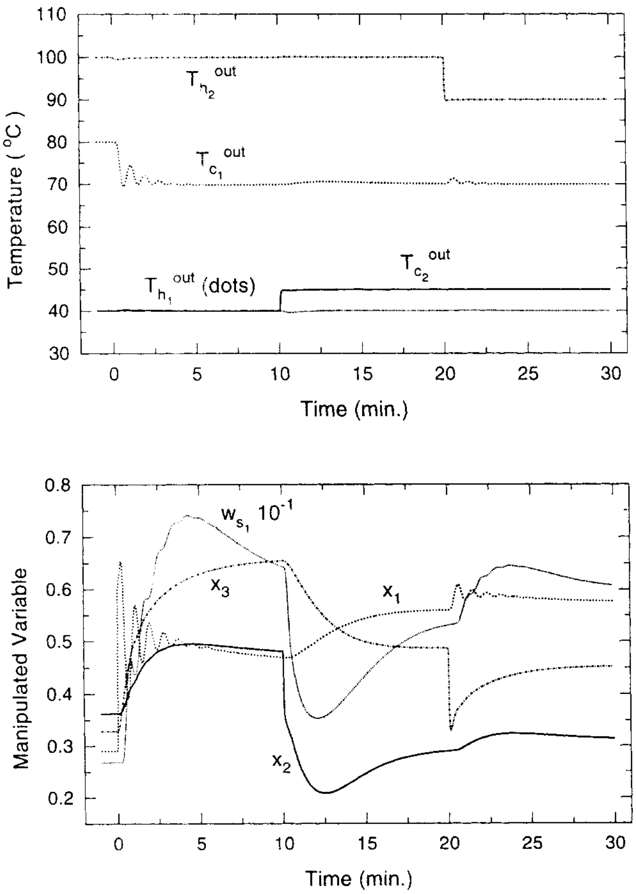

Figure 6. Dynamic responses of *design A* to the sequence of temperature setpoint changes.

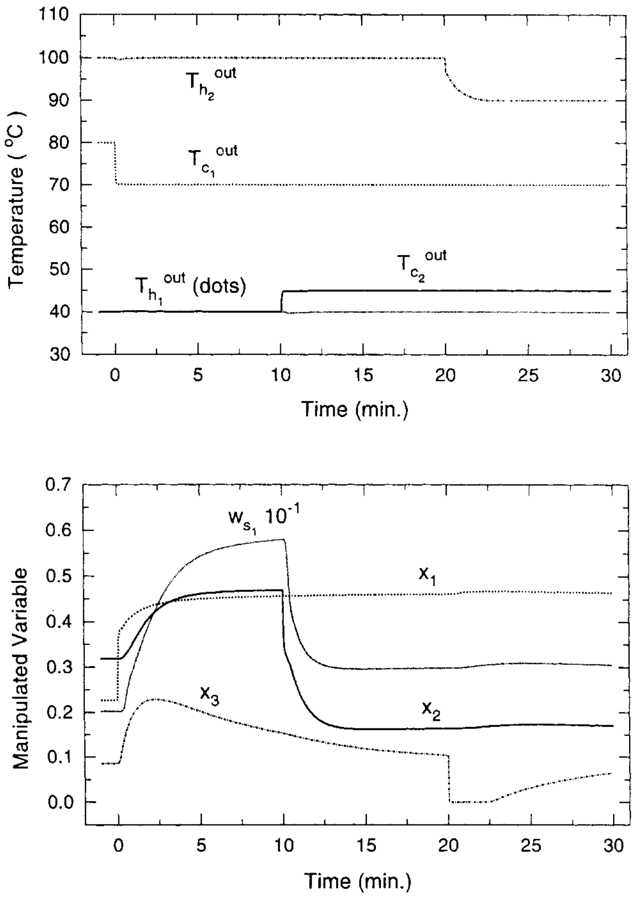

Figure 7. Dynamic responses of *design B* to the sequence of temperature setpoint changes.

$$e_k(F_T) = \frac{\left(T_j - T_j^0\right)}{\left(T_i^0 - T_j^0\right)}, \tag{40}$$

for all equipment, k, having constraints in the HEN model.

The right side of Eq. 40 is the actual efficiency value computed, for instance, at the nominal operating point, based on actual temperatures associated with each plant heat-exchanger unit. Table 2 shows F_T values determined in this manner for our example problem, where the nonlinear simulator replaced the actual plant.

The sequence of setpoint changes is as follows: first, at time $t = 0$, the target for $T_{c_1}^{out}$ changes from 80°C to 70°C; ten minutes later the setpoint for $T_{c_2}^{out}$ changes from 40°C to 45°C, and after another ten minutes $T_{h_2}^{out}$ is taken to 90°C from 100°C. Now, the optimizer computes the new commands at virtually the same time that setpoint changes are made. Tables 3 and 4 show that the amount of energy integration achieved in this sequence is the same for both designs, but again *design B* needs the protection activated for x_3 in two of these steady states, something that is not necessary for *design A*.

Figures 6 and 7 show the dynamic responses to the setpoint changes of controlled and manipulated variables. In Figure 6, x_1 seems too oscillatory when changing the setpoint for $T_{c_1}^{out}$; however, reducing the controller sensitivity deteriorates responses to load changes. Figure 7 also shows that after the third setpoint change is made, the bypass to exchanger 3 remains temporarily closed, reducing control performance on $T_{h_2}^{out}$; this happens because the cooler S_2, which must go from a duty of 169.5 kW to the required 369.5 kW, has a much larger lag than the direct acting bypass. Recall that the optimizer works with steady-state conditions only, and since optimization leads the system to operate with low bypass ratios, we must be aware that temporary saturation might occur. However, since most closed loops are faster than internal HEN dynamics, most saturations occurring immediately after the change can be anticipated.

Remark 11 (Temporary Saturations). Let us call Δq_k the change of duty that a given unit $k \in \{1, ne + s\}$ in a network must perform as part of a set of actions to drive the system to a new operating point, and q_k to the actual duty. Then, temporary saturation of the associated manipulated variable might occur if

$$|\Delta q_k| > q_k \qquad \text{or} \qquad |\Delta q_k| > e_k^o L_k^o - q_k.$$

This means that comparing slack variables at the actual steady state with the required change might anticipate eventual temporary saturations during a shift in the operating point. The problem also suggests the implementation of a what-if parallel task at the supervisor level. As a matter of fact, *once the coordination between levels is adjusted, the supervisor works not only optimizing, but it also protects regulation.*

The analysis made in this section, though using a single example, exposes benefits and limitations of HEN optimization using constant control configuration. Several other examples were successfully analyzed, modeled, and configured, including networks with combined split streams or with nonconvex operating spaces. Results and comments about these examples are not included here due to space limitations. Good control performance is achieved using the proposed procedure. Of course, other approaches can be taken; in particular, a decentralized control system with varying configuration based on control logics (Mathisen et al., 1994). A comparison between these procedures is an interesting work to be undertaken in the future.

Conclusions

This article shows a simple and proper modeling procedure for on-line optimization of HENs, provides guidelines to define the loop configuration of the regulating system, and reveals the main clues for coordinating with the supervisor level.

The modeling approach used in the optimization problem emphasizes the coordination between the supervisor and the regulating system by providing clear links with the constraints of manipulated variables. The optimization problem assumes that the network structure, the heat exchanger areas, the stream targets, and inlet conditions are known; the objective function is formulated in terms of the minimization of utility consumptions or the maximization of the energy integration. The optimal network condition is determined by solving an LP or NLP problem, depending on the complexity of the network, the control objectives, and the working space under which the operation is required.

Modeling equations presented in this article are useful not only for on-line optimization of a HEN operation, but they also serve to have preliminary solutions to the optimization problem that can be used to provide information for configuring the regulating system. The slack variables associated with the individual unit constraints help in determining unnecessary bypasses, identifying permanent and temporary units, and coordinating actions between the regulating system and the optimizer so as to avoid permanent saturation of manipulated variables.

Several additional features are highlighted by the simulation experiences; (1) the problem formulation for an on-line optimizer must denote most preferences and decisions adopted during the definition of the associate control configuration; (2) correction factors to account for deviations from pure countercurrent pattern help to reduce the mismatch between the HEN model used by the optimizer and the actual plant; (3) inspection of slack variables prior to an operation shift helps to determine whether a temporary saturation of the manipulated variables might occur immediately after the change is made; (4) a what-if parallel task at the supervisor level is also advisable.

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Literature Cited

Aguilera, N., and J. L. Marchetti, "Supervisory Control of Heat Exchanger Networks," *Proc. Symp. on Low Cost Automation*, AADECA-IFAC, Buenos Aires, Argentina, p. 73 (1995).
Beautyman, A. C., and A. R. H. Cornish, "The Design of Flexible Heat Exchanger Networks," *Proc. U.K. Nat. Heat Transfer Conf.*, IChemE Symp. Ser. (No. 86), Vol. 1, p. 547 (1984).
Calandranis, J., and G. Stephanopoulos, "Structural Operability Analysis of Heat Exchanger Networks," *Chem. Eng. Res. Des.*, **64**, 347 (1986).
Calandranis, J., and G. Stephanopoulos, "A Structural Approach to the Design of Control Systems in Heat Exchanger Networks," *Comput. Chem. Eng.*, **12**(7), 651 (1988).
Colberg, R. D., and M. Morari, "Area and Capital Cost Targets for Heat Exchanger Networks Synthesis with Constrained Matches and Unequal Heat Transfer Coefficients," *Comput. Chem. Eng.*, **14**(1), 1 (1990).
Correa, D. J., "Dinámica y Control de Sistemas de Intercambio Calórico," Masters Thesis, Univ. Nacional del Litoral, Santa Fe, Argentina (1994).
Correa, D. J., and J. L. Marchetti, "Dynamic Simulation of Shell-and-Tube Heat Exchangers," *Heat Transfer Eng.*, **8**(1), 50 (1987).
Dodge, B. F., *Chemical Engineering Thermodynamics*, McGraw-Hill, New York (1994).
Floudas, C. A., and I. E. Grossmann, "Synthesis of Flexible Heat Exchanger Networks for Multiperiod Operation," *Comput. Chem. Eng.*, **10**(2), 153 (1986).
Galli, M. R., and J. Cerdá, "Synthesis of Flexible Heat Exchanger Networks: III. Temperature and Flowrate Variations," *Comput. Chem. Eng.*, **15**(1), 7 (1991).
Huang, Y. L., and L. T. Fan, "Distributed Strategy for Integration of Process Design and Control," *Comput. Chem. Eng.*, **16**(5), 497 (1992).
Kern, D. Q., *Process Heat Transfer*, McGraw-Hill, New York (1950).
Kotjabasakis, E., and B. Linnhoff, "Sensitivity Tables in the Design of Flexible Processes," *Chem. Eng. Res. Des.*, **64**, 197 (1986).
Marselle, D. F., M. Morari, and D. F. Rudd, "Design of Resilient Processing Plants: II. Design and Control of Energy Management System," *Chem. Eng. Sci.*, **37**(2), 259 (1982).
Mathisen, K. W., "Integrated Design and Control of Heat Exchanger Networks," PhD Thesis, Univ. of Trondheim, Trondheim, Norway (1994).
Mathisen, K. W., S. Skogestad, and E. A. Wolf, "Bypass Selection

for Control of Heat Exchanger Networks," Eur. Symp. on Computer Aided Process Engineering—1, Elsinove, Denmark (1992).
Mathisen, K. W., M. Morari, and S. Skogestad, "Optimal Operation of Heat Exchanger Networks," PSE'94, Kyungju, Korea (1994).
Papalexandri, K. P., and E. N. Pistikopoulos, "A Multiperiod MINLP Model for Improving the Flexibility of Heat Exchanger Networks," Eur. Symp. on Computer Aided Process Engineering—2, Toulouse, France (1992).
Rotea, M. A., and J. L. Marchetti, "Integral Control of Heat-Exchanger-Plus-Bypass Systems," IEEE Int. Conf. on Control Applications, Hartford, CT (1997).
Shinskey, F. G., *Distillation Control for Productivity and Energy Conservation*, McGraw-Hill, New York (1977).
Yee, T. F., and I. E. Grossmann, "Simultaneous Optimization Models for Heat Integration: II Heat Exchanger Network Synthesis," *Comput. Chem. Eng.*, **14**(10), 1165 (1990).

Appendix A: About Degrees of Freedom and Structural Network Flexibility

We refer to the *structural flexibility* of a HEN, when it depends only on the number of process streams with and without a final target, the number of process–process exchangers, the amount of service equipment, and how all these elements are connected. The concept does not include unit constraints arising from limited exchange areas or stream-match conditions as in Eq. 17. Our claim is that the number (9) that counts the *independent variables in the total energy balance* gives a new insight about structural flexibility. The meanings of different f values are discussed next.

1. $f = -1$, means that there are no utility units and that all the process streams have temperature targets. It does not matter how the structure is defined, the network cannot be operated since there is no way out to any eventual energy disturbance. There is no compatibility between the proposed network structure and the number of desired targets for the regulating system. The system is not controllable.

2. $f = 0$, there is one outlet-free stream—utility or process stream—that allows the operation of the network. The control loop using this outlet-free stream as manipulated cannot be protected from saturation. Once the HEN regulating system achieves the control targets, the energy integration is completely determined. This kind of network might have flexibility to vary excess bypasses or split ratios in limited ranges while satisfying all the regulating targets, but the amount of energy integration for the overall network would not be improved. These multiple but equivalent operating points imply a flexibility that does not allow optimization, but might help to protect manipulated bypasses from saturation.

3. $f > 0$, there is more than one stream path to release energy disturbances. We may have many steady states with different utility loads—for the same targets and input conditions—and consequently optimization makes sense. However, three subcases must be noted:

a. All the stream paths that allow the release of energy disturbances are associated with $\mathcal{H}_2$ and $\mathcal{H}_3$. This means that all the utility units are for cooling operations only, and eventually there are hot process streams without temperature targets. If these utilities have the same cost, and there are no free-outlet streams (which is equivalent to a null cost utility), then the total utility load cannot be affected. On the other hand, if the cooling utilities have different costs, then an optimizing system would be advisable.

b. All the stream paths that allow the release of energy disturbances are associated with $\mathcal{C}_2$ and $\mathcal{C}_3$. This means that all the utility units are for heating operations only, and eventually there are cold process streams without temperature targets. Similar to case (a), an optimizing supervisor would be advisable if the heating utilities have different costs or if a free-outlet cold process stream is available, or both.

c. The network has heating and cooling utilities available and eventually, process streams without temperature targets. This is the case for which an optimizing supervisor control could yield immediate benefits, since there exist enough degrees of freedom to search for an optimal operating point.

Now, let us analyze the structures in Figure A1 to visualize the information provided by the *degrees of freedom of the total energy balance* in Eq. 9. For the sake of simplicity, let us assume that all the process streams have temperature targets and all the process–process exchangers have bypasses. Let us also consider the degrees of freedom suggested by Marselle et al. (1982): $f_m = ne + s - n_{\mathcal{H} \cup \mathcal{C}}$, to see the differences.

Case a. $f = -1$, though there are four bypasses and four targets to be controlled, this structure is not controllable: the lefthand side in Eq. 8 is zero, so there is no escape path for energy disturbances—a change of inlet temperature in the righthand side, for instance—and at least one stream target must be resigned. Compensating disturbances are disregarded for practical reasons. On the other hand, $f_m = 0$ means that the system cannot be optimized but it does not prevent considering controllability.

Case b: f = 1. Observe that within certain ranges, the utility duties can be varied by changing the load q_1. The system is controllable and may be optimized. The action for optimization is trivial in this case, since the largest load q_1 will yield the smallest utility duties, but the example is useful to see the meaning of $f > 0$: utility duties can be different for

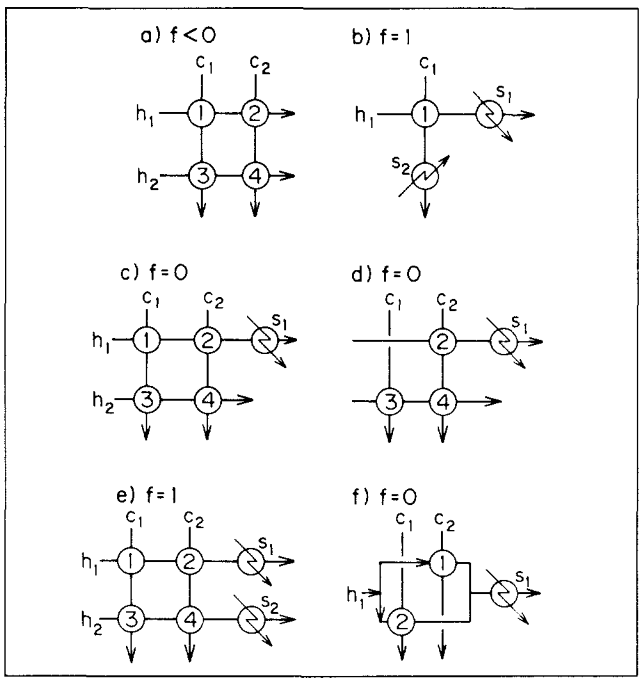

Figure A1. Number of independent outlet-free streams in different net diagrams.

the same targets and inlet conditions. The fact that $f_m = 1$ is just a coincidence: $f_m > 0$ simply means that there are more units than process streams, which is clear in the next case.

Case c: $f = 0$ and $f_m = 1$. The first number says that the network is controllable but optimization is not possible: it does not matter which one is the load q_1, once the control targets are satisfied this service duty is determined and cannot be changed; there is no way to improve energy integration. The second number says that q_1 exists and that it provides the flexibility to protect, within certain ranges, manipulated bypasses from saturations; however, the control loop associated with the service unit cannot be protected, because this service cannot be constrained at all. The system is controllable and might need a supervisor (or logic, or split-range controllers, etc.) to care for saturations so the full flexibility structurally available can be used.

Case d: $f = 0$ and $f_m = 0$. The interpretation of the numbers in this case is quite clear, as compared to case c. The system is controllable, cannot be optimized, and has less flexibility than in the previous case since no manipulated variable can be protected. In comparison with case a, it can be observed that f_m does not differentiate between them since it gives zero in both cases.

Case e. As in case b, $f = 1$ means that there are two escape paths for energy disturbances; however, in that case when one utility load increases, the other does the same. In this case, we have the freedom to balance the heat loads in order to emphasize the use of one utility or the other, but the total cooling load cannot be changed; observe these conditions in Eq. 8. This flexibility to move the system around different steady states can be used to protect control loops from saturation or to optimize in case of different utility costs. On the other hand, $f_m = 2$ means that there are two shift variables (Marselle et al., 1982) to move the system around, but it might not mean optimization at all—in the sense of possibilities for minimizing total utilities.

Case f: $f = 0$ and $f_m = 0$. As in the previous cases, where the first number is zero, it means that heat integration cannot be improved and the control loop associated with the service unit cannot be protected. The hot-stream split ratio is also a shift variable that incorporates, in certain ranges, flexibility for protecting bypasses from saturation; however, the second number does tell about that flexibility in this case.

Since f is supported by Eq. 8, a good deal of insight is gained about structural flexibility for optimization, in the sense of having possibilities to change a performance index like Eq. 18 or Eq. 22. On the other hand, once the target control system is defined, $f_m + n_{\text{split}}$ is the number of variables that remain to be fixed by the supervisor just to take care of control saturations, optimization, or both.

Appendix B: The Heat Exchanger Efficiency and Related Parameters

Equation B1, where 0 stands for inlet conditions, shows that heat duty q can be written in terms of energy released by the hot stream, i, or the energy taken by the cold stream, j, or the heat-transfer equation where A is the heat-exchanger area, U the overall heat-transfer coefficient, and F_T a factor correcting ΔT_{ml} to account for deviations from the pure countercurrent pattern (Kern, 1950),

$$q = w_i c_i (T_i^0 - T_i) = w_j c_j \left(T_j - T_j^0\right) = UA\{F_T \Delta T_{ml}\}. \quad \text{(B1)}$$

The second equality leads to defining the heat-capacity flow-rate ratio as

$$R = \frac{w_j c_j}{w_i c_i} = \frac{(T_i^0 - T_i)}{\left(T_j - T_j^0\right)}, \quad \text{(B2)}$$

algebraic rearrangements combining the third equality show the convenience of defining the number of transfer units as

$$N_{TU} = \frac{UA}{w_j c_j}, \quad \text{(B3)}$$

and, the exponential relationship,

$$B = \exp\{N_{TU} F_T (R - 1)\} = \frac{\left(T_i^0 - T_j\right)}{\left(T_i - T_j^0\right)}. \quad \text{(B4)}$$

The preceding parameters help the computation of the heat-exchanger efficiency defined by Dodge (1944),

$$e = \frac{1 - B}{1 - RB} = \frac{\left(T_j - T_j^0\right)}{\left(T_i^0 - T_j^0\right)} \quad \text{for} \quad R \neq B \neq 1. \quad \text{(B5)}$$

Observe, from the definition (Eq. B4), that if $R \to 1$, then $B \to 1$ also. The application of limits through the L'Hospital rule to (Eq. B5) allows computing e for this condition,

$$e = \frac{N_{TU}}{1 + N_{TU}} \quad \text{for} \quad R = B = 1. \quad \text{(B6)}$$

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